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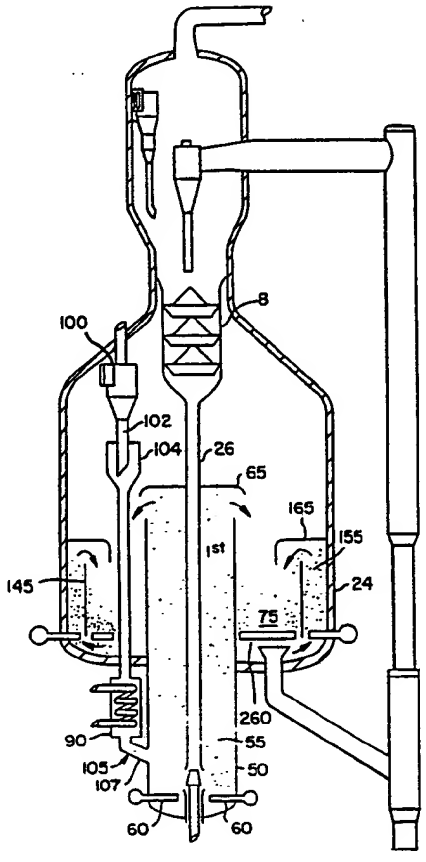
(54) Title: PROCESS FOR REGENERATING FLUIDIZED CATALYTIC CRACKING CATALYST

(57) Abstract

A process is disclosed for achieving turbulent or fast fluidized bed regeneration of spent FCC catalyst in a bubbling bed regenerator (24) having a stripper (8) mounted over the regenerator and a stripped catalyst standpipe (26) within the regenerator (24). A coke combustor vessel (50), which may be partially or totally open to the dilute phase above the bubbling bed (75), is added to the existing regenerator vessel. Spent catalyst is discharged into the coke combustor (50), regenerated in a turbulent or fast fluidized bed (55), then discharged into the dilute phase region above the bubbling bed (75). Regeneration of catalyst is completed in the bubbling dense bed (75), and/or an annular fast fluidized bed (155) surrounding the coke combustor (50). Catalyst may be recycled from the dense bed (75) to the coke combustor (50).



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PROCESS FOR REGENERATING FLUIDIZED CATALYTIC CRACKING CATALYST

The invention relates to a process for regenerating fluidized catalytic cracking catalyst.

In the fluidized catalytic cracking (FCC) process, catalyst circulates between a cracking reactor and a catalyst regenerator. In the reactor, hydrocarbon feed contacts a source of hot, regenerated catalyst, which vaporizes and cracks the feed at a temperature fo 425-600°C, usually 460-560°C. The cracking reaction deposits carbonaceous hydrocarbons or coke on the catalyst, thereby deactivating the catalyst. The cracked products are separated from the coked catalyst, which is then stripped of volatiles, usually with steam, in a catalyst stripper. The stripped catalyst is then passed to the catalyst regenerator, where coke is burned from the catalyst with oxygen containing gas, usually air. Decoking restores catalyst activity and simultaneously heats the catalyst to, for example, 500-900°C, usually 600-750°C. This heated catalyst is recycled to the cracking reactor to crack more fresh feed. Flue gas formed by burning coke in the regenerator may be treated for removal of particulates and for conversion of carbon monoxide, after which the flue gas is normally discharged into the atmosphere.

Catalytic cracking has undergone progressive development since the 1940's. The trend of development of the fluid catalytic cracking (FCC) process has been to all riser cracking and the use of zeolite catalysts. A good overview of the importance of the FCC process, and its continuous advancement, is provided in "Fluid Catalytic Cracking Report", by Amos A. Avidan, Michael Edwards and Hartley Owen, published in the January 8, 1990 edition of the Oil & Gas Journal. One modern,

compact FCC design is the Kellogg Ultra Orthoflow converter, Model F, which is shown in Figure 1 of the accompanying drawings and also shown as Figure 17 of the January 8, 1990 Oil & Gas Journal article discussed above. Although this unit works well in practice, its use of a bubbling bed regenerator is inherently inefficient, and troubled by the presence of large bubbles, poor catalyst circulation, and the presence of stagnant regions. Moreover, bubbling bed regenerators usually have much larger catalyst inventories and longer catalyst residence times to make up for a lack of efficiency. For such units, characterized by a stripper mounted over, and partially supported by, the bubbling dense bed regenerator, there has been no good way to achieve the benefits of high efficiency regeneration.

The present invention seeks to provide a fluidized catalytic cracking process which exhibits the compact design of the Kellogg unit described above but which achieves improved catalyst regeneration.

Accordingly, the present invention resides in a fluidized catalytic cracking process wherein a heavy hydrocarbon feed is cracked to lighter products comprising the steps of:

catalytically cracking said feed in a riser reactor by mixing the feed in the base of the reactor with a source of hot regenerated catalytic cracking catalyst withdrawn from a catalyst regenerator, and cracking said feed in said riser reactor to produce catalytically cracked products and spent catalyst which are discharged from the top of the riser into a catalyst disengaging zone;

separating cracked products from spent catalyst in said catalyst disengaging zone to produce a cracked product vapor phase which is recovered as a product and a spent catalyst phase which is discharged from said

disengaging zone into a catalyst stripping zone contiguous with and beneath said disengaging zone;

stripping said spent catalyst with a stripping gas in said stripping zone to produce a stripper vapor comprising cracked products and stripped catalyst, which is discharged into a vertical standpipe beneath said stripping zone;

discharging stripped catalyst from said standpipe into a catalyst regenerator and regenerating said stripped catalyst in said regenerator, said regenerator comprising a first regeneration zone which is located beneath said stripping zone and which comprises a single dense phase bubbling fluidized bed of catalyst to which an oxygen containing regeneration gas is added and from which hot regenerated catalyst is withdrawn and recycled to said riser reactor, characterized in that:

said regenerator comprises a second, coke combustion zone which is at least partially immersed in the bubbling fluidized bed of the first regeneration zone and into which stripped catalyst is discharged from said standpipe;

an oxygen-containing gas is added to said coke combustion zone partially to regenerate the stripped catalyst and maintain a majority of the catalyst in a state of turbulent or fast fluidization; and

the partially regenerated catalyst and flue gas is discharged from said coke combustion zone to a dilute phase region of the first regeneration zone above the bubbling fluidized bed and at least part of the catalyst is separated from the flue gas and collected in said bubbling fluidized bed for further regeneration.

The invention will now be more particularly described in the accompanying drawings, in which

Figure 1 is a schematic view of a conventional fluidized catalytic cracking unit,

Figure 2 is a schematic view of a regenerator according to a first example of the invention,

Figure 3 is a schematic view of a modification of the first example,

Figure 4 is a schematic view of a regenerator according to a second example of the invention,

Figure 5 is a schematic view of a regenerator according to a third example of the invention, and

Figure 6 is a schematic view of a regenerator according to a modification of the third example.

Referring to the drawings, Figure 1 is a simplified schematic view of an FCC unit of the prior art, similar to the Kellogg Ultra Orthoflow converter Model F shown as Fig. 17 of Fluid Catalytic Cracking Report, in the January 8, 1990 edition of Oil & Gas Journal.

A heavy feed such as a vacuum gas oil is added to the base of the riser reactor 6 via feed injection nozzles 2. The cracking reaction is completed in the riser reactor and spent catalyst and cracked products are discharged by way of 90° elbow 10 to riser cyclones 12. The cyclones 12 separate most of the spent catalyst from cracked product, with the latter being discharged into disengager 14, and eventually removed via upper cyclones 16 and conduit 18 to a fractionator (not shown).

Spent catalyst is discharged from a dipleg of riser cyclones 12 down into catalyst stripper 8, where one, or preferably 2 or more, stages of steam stripping occur, with stripping steam admitted by means not shown in Figure 1. The stripped hydrocarbons, and stripping steam, pass into disengager 14 and are removed with cracked products after passage through upper cyclones 16.

Stripped catalyst is discharged down via spent catalyst standpipe 26 into catalyst regenerator 24,

with the flow of catalyst being controlled by a spent catalyst plug valve 36.

Catalyst is regenerated in regenerator 24 by contact with air, added via air lines and an air grid distributor (not shown). A catalyst cooler 28 is provided so that heat may be removed from the regenerator, if desired. Regenerated catalyst is withdrawn from the regenerator via regenerated catalyst plug valve assembly 30 and fed via lateral 32 into the base of the riser reactor 6 to contact and crack fresh feed injected via injectors 2, as previously discussed. Flue gas, and some entrained catalyst, are discharged into a dilute phase region in the upper portion of regenerator 24. Entrained catalyst is separated from flue gas in multiple stages of cyclones 4, and the separated flue gas is collected in a plenum 20 for discharge to a flare via line 22.

In Figure 2 only the differences from the regenerator 24 of Figure 1 are shown. Like elements in Figure 1 and 2 have like numerals.

Thus, referring to Figure 2, in the example shown, a vertically extending, cylindrical coke combustor 50 is added to and passes through the base of the regenerator vessel 24. The coke combustor 50 has its lower end closed and its upper end connected to the stripper 8 and the remainder of the regenerator vessel 24. Stripped catalyst from the catalyst stripper is discharged via stripper dipleg 26 down into the coke combustor 50 to form a relatively dense, fast fluidized bed (FFB) region 55, where incoming spent catalyst contacts regeneration gas, usually air, added via multiple inlets 60.

The partially regenerated catalyst, and partially consumed combustion gas are discharged from the vessel 50 via a bubble cap 65 which extends over the open end of the vessel 50 and deflects the mixture downwardly into the much larger volume inside the vessel 24. The

rapid increase in volume, or in cross sectional area available for fluid flow, results in a rough but rapid separation of catalyst from flue gas. A majority, preferably over 90 wt% of the catalyst is discharged downwardly in a relatively compact mass toward the dense bed of catalyst 75 in the base of the existing regenerator shell 24. Air is added to bed 75 via air ring 160 to maintain fluidization and preferably to achieve a significant amount of coke combustion. Although bed 75 is a typical fluidized bubbling bed, characterized by relatively large stagnant regions, and large bubbles of combustion air which bypass the bed, it is an excellent place to achieve some additional coke combustion. One of the most significant benefits of coke combustion in bubbling bed 75 is the relatively drier atmosphere. There is a lower steam partial pressure in the dense bed 75 of the present invention than in a conventional dense bed regenerator, such as that shown in Fig. 1. Much of the reduction in steam partial pressure is due to the removal of water of combustion, and entrained stripping steam, with the flue gas discharged from the coke combustor. By using a flue gas/catalyst separation means such as cap 65 on the transport riser outlet, the relatively high steam content flue gas is separated from the catalyst before the latter is discharged down to form the bubbling fluidized bed 75. It is also possible to reduce the load on the cyclones 100 above the bubbling dense bed, because much less combustion air is needed, and consequently there is less entrainment of catalyst into the dilute phase, when only a fraction of the coke combustion occurs in the bubbling dense bed. Even without a separation means such as cap 65, the dense bed region 75 of the present invention will be drier than the dense bed of the regenerator of Figure 1.

In the preferred embodiment shown, an additional stage of combustion is provided by annular region 155

defined by a baffle 145 and the walls of regenerator vessel 24. Catalyst from dense bed region 75 flows under baffle 145, contacts additional combustion air added via air ring 260 and flows up into the region 155. Preferably enough air is added, relative to the cross sectional area, to result in superficial vapor velocities which produce a turbulent fluid bed or more preferably a fast fluidized bed. In this way additional coke combustion, and afterburning of CO to CO_2 can be achieved in an efficiently fluidized bed, which is extremely dry. The coke on catalyst will have a very low hydrogen content, because all of the "fast coke" will have been burned in the coke combustor, and a majority of the hydrogen content of the remaining coke will be eliminated in the dense bed 75. Coke combustion in the third stage region 155 will be free of the two major sources of steam in FCC regenerators, namely water of combustion and entrained stripping steam. Thoroughly regenerated catalyst is discharged from the top of region 155 via radial deflector 165, which functions much like bubble cap 65 in that a significant separation of catalyst from flue gas is achieved.

Preferably from 20 to 90 wt% of the coke combustion occurs in the coke combustor and dilute phase transport riser. A further 5 to 50 wt%, and most preferably 10 to 40 wt%, of the coke combustion occurs in the bubbling bed 75. Another 5 to 50 wt%, and most preferably 10 to 40 wt%, of the coke combustion occurs in the third stage combustion zone 155.

The optimum amount of coke combustion that occurs in each zone will depend on a number of factors, including the amount of sulfur and nitrogen in the feed, rate of catalyst replacement, and metals contamination of the feed. For cleanest catalyst, when metals and NOx emissions are not a problem, it is beneficial to front load the air addition, i.e., to

maximize coke combustion in bed 55. To minimize NOx, coke combustion should be delayed, so that large amounts of carbon will be present to hinder NOx formation.

It is possible, by means not shown in Figure 2, to divert catalyst discharged from the third stage region 155 to a catalyst reservoir supplying hot regenerated catalyst for recycle to the reactor via line 32. This minimizes backmixing of catalyst from the third stage region 155 with catalyst in the bubbling dense bed region 75.

It will be frequently be beneficial to recycle some hot regenerated catalyst to the fast fluidized bed region in coke combustor 50. Such recycle can come from the dense bed 75, or preferably, from a primary cyclone such as cyclone 100, as shown in the drawing. Hot catalyst is discharged down dipleg 102 into a catalyst return funnel 104, which can be much higher than the top of the bubbling dense bed 75. Accordingly, a large head will be available to permit controlled transfer of hot regenerated catalyst from the cyclone dipleg to the coke combustor 50, with flow control achieved via slide valve 105. Regenerated catalyst is then charged to the region 50 via line 107.

A catalyst cooler 90 may be provided to remove some heat from the regenerator, if heavy crudes or unusual operating conditions prevent a classical heat balanced operation. Catalyst coolers can also be associated with the dense bed 75, the region 50, or on the return line to the reactor.

Figure 3 shows a modification of the Figure 2 example in which the bubble cap 65 is omitted. Mechanically, this is the easiest way to achieve the benefits of fast fluidized bed coke combustion, at minimum capital cost. The Figure 3 embodiment even allows a significant amount of catalyst recycle, i.e., recycle of hot regenerated catalyst from the bubbling

dense bed to the coke combustor, without a catalyst recirculation line or valve. Catalyst recycle can be achieved by regulating the relative depths of the dense bed 75 and the sidewalls of the coke combustor 50. Operation with a relatively high dense bed 75 level will result in considerable circulation of hot regenerated catalyst into coke combustor 50. Lowering of the dense bed 75 level, such as by reducing the superficial vapor velocity in the bed, or operating with a lower catalyst inventory, will reduce the tendency of hot regenerated catalyst from bed 75 to overflow into the coke combustor 50. The Figure 3 approach will achieve a relatively drier regeneration in bed 75, because any steam discharged from the coke combustor will tend to travel up.

A drawback to the approach shown in Figure 3 is that there can be some increase in catalyst traffic in the dilute phase region 70 above bubbling dense bed 75, especially when a large amount of coke combustion occurs in this region. This can be tolerated in many units, because the amount of combustion air needed, and the resulting superficial vapor velocity, in bubbling bed 75 can be greatly reduced or eliminated. There will be a large increase in catalyst traffic near the open end of the coke combustor 50, but this will be partially or totally offset by a reduction in catalyst traffic above bubbling bed 75. Where desired, improved cyclones, precipitators, or other conventional means may be added to permit more catalyst entrainment in the dilute phase above bubbling dense bed 75.

Figure 4 illustrates a second example of the invention in which a bell-shaped coke combustor vessel 250 is added to, and contained within, the existing regenerator vessel 24. The vessel 250 includes a wide base region 240 which is open to and interacts with the existing bubbling dense bed 75 to create a localized

region 260 of turbulent fluidization, or fast fluidization.

Stripped catalyst from the catalyst stripper 8 is discharged via stripper dipleg 26 down into the regenerator to meet plug valve 36. The catalyst is discharged into the vessel 250 to form the relatively dense, fast fluidized bed (FFB) region 260, where incoming spent catalyst contacts regeneration gas, usually air, added via multiple inlets 238. In bell shaped vessel 250 the air admission rate, the cross-sectional area available for flow at the base region 240, and catalyst addition and catalyst recycle, are adjusted to maintain much or all of the bed in a "fast fluidized condition", characterized by intense agitation, relatively small bubbles, and rapid coke combustion. The superficial vapor velocity and catalyst density required to achieve this condition are described below, but in the upper regions of the coke combustor, where the diameter of the bell shaped vessel decreases, as indicated generally at 242, and flow approaches dilute phase conditions, the catalyst density the density will be 0.2 gm/cc (10 pounds/cubic foot), or less.

The partially regenerated catalyst, and partially consumed combustion gas are discharged through the top of FFB region 260 into a dilute phase transport riser 244, which forms an annular region 265 around the spent catalyst standpipe 26. Dilute phase conditions prevail in the region 265 here rapid combustion of CO to CO2, and some additional coke combustion can also be achieved. Addition of secondary air, to the base of the transport riser 244, or at higher elevations therein by means not shown, can also be practiced to augment coke or CO combustion.

The catalyst and flue gas are discharged into the dilute phase space 70 above the existing bubbling dense bed 75. A catalyst/flue gas separation means, such as

cyclones 248, connected to the riser 244 by transfer lines 246, are used to separate the bulk of the catalyst from the bulk of the flue gas, and reduce the amount of catalyst dispersed in the dilute phase region 70. Hot regenerated catalyst recovered by the cyclones 248 is discharged down via diplegs 254 into the bubbling dense phase fluidized bed 75 in the base of the existing regenerator shell 24. Additional regeneration gas and/or fluffing air to maintain fluidization may be added to the bed 75 if desired.

It will be beneficial to recycle some hot regenerated catalyst from bubbling dense bed 75 into the fast fluidized bed region 260 in bell shaped vessel 252. Such recycle can come from flow under the bell lower portion 240, from the dipleg 254 of a cyclone, or via an aspirated lift tube. Catalyst underflow is preferred because it is simple, reliable and controllable. Catalyst underflow can be assisted by horizontal air jets 88, or vertical jets (not shown) in the FFB region near the base section 40 of the bell shaped vessel 250. There are no slide or plug valves, and no narrow passageways or pipes, needed to transfer catalyst from the bubbling dense bed to the FFB region. Control of catalyst recycle can be achieved by increasing the level of catalyst in the bubbling dense bed 75, by decreasing the pressure in the base of the FFB region 260, or by conventional aspiration techniques, such as the air jets 288.

The regenerator shown in Figure 4 has a very large operating window in regards to catalyst recirculation rate. The unit will operate well with very low rates of catalyst recirculation, because of the efficient nature of FFB catalyst regeneration, and because the incoming spent catalyst will be preheated to a significant extent in the stripper standpipe, especially when the dilute phase transport riser encompasses the standpipe as shown in Figure 4. The

unit tolerates operation with large amounts of recycle, the catalyst merely circulates rapidly within the regenerator vessel. The use a catalyst/flue gas separation device, such as the cyclones 248, is preferred so that even if large catalyst circulation rates are achieved the catalyst traffic in the dilute phase region 70 above the dense bed 75 will not be inordinately increased. Operation with large amounts of recycle reduces the apparent coke concentration of catalyst in the FFB region, but increases the temperature and the coke burning rate so efficient regeneration is still achieved, even with 5:1 or higher recycled:spent catalyst ratios.

The bell shaped vessel 250 is preferred for its efficiency, i.e., it achieves an efficient and gradual increase in superficial vapor velocity within the FFB region, providing a smooth transition to dilute phase flow in or near the inlet to the transport riser. It also presents a large sloping surface to the dilute phase region within the regenerator dilute phase, which promotes catalyst settling and gradual compaction to form a dense phase bubbling bed in region 270. Such a shape provides a gradual response to changes in air or spent catalyst flow so that any change in a process flow causes only a minor change in the operation of the FFB region or the bubbling bed region.

An alternative to the bell-shaped design is an inverted cone which, despite some degradation in performance, may be desirable because of cheaper fabrication and/or installation costs.

Also suitable, but not preferred, are designs with stepped changes in diameter. An inexpensive, but hard to control, design calls for two lengths of pipe, with different diameters. For example,if the regenerator vessel 24 had a diameter of 10 meters, a FFB region could be formed by suspending a suitable length, say 2-10 meters, of pipe having a diameter of 6-9.5 meters,

and preferably of 7-9 meters, from the top of the regenerator. The dilute phase transport riser could comprise another suitable length of pipe, say 3-15 meters, having a diameter of 1-4 meters, and preferably 1.5-3.5 meters, and sufficient to accommodate the stripper standpipe 26. This could also be supported from the top of the regenerator, or from the stripper. The two diameters of pipe could then be connected via an annular weld cap. Such an arrangement, because of its sudden changes in diameter, could respond very differently to small changes in process flows, so greater care would be needed to assure smooth operation.

The coke combustors shown in Figures 2-4 can benefit significantly from indirect heat exchange, i.e., the transfer of heat from the bubbling dense bed 75 into coke combustor 50. Use of relatively conductive, rather than insulating, refractory linings, heat pipes, fins, dimples, and the like can be used to increase indirect heat exchange from bed 75 into the coke combustor. Indirect heat exchange is highly beneficial in reducing catalyst traffic in the coke combustor, and hence catalyst carry-over into the dilute phase region, and in reducing exposure of regenerated catalyst to the relatively high steam partial pressures which occur in the coke combustor due to water of combustion.

Referring to Figure 5, in the third example, a bottle-shaped coke combustor vessel 350 is added to, and extends through, the base of the existing regenerator 24. The vessel is closed at the lower end of its wide body portion 351 and tapers over a frusto-conical portion 354 to a narrow neck portion 356 located within the dilute phase region 70 of the regenerator.

The standpipe 26 extends down into body portion 351 of the vessel 350 and discharges stripped catalyst

into a fast fluidized bed region 352, where incoming spent catalyst contacts combustion gas, usually air, added via multiple inlets 360. The partially regenerated catalyst, and partially consumed combustion gas are discharged through the portions 354, 356 out the top of the coke combustor into the dilute phase region 70 of the regenerator 24. Preferably a catalyst/flue gas separation means, such as the bubble cap 358, is used to separate the bulk of the catalyst form the bulk of the flue gas, and reduce the catalyst traffic in the dilute phase region 70. Alternatively, cyclones can be used and can easily achieve greater than 99 % separation, but they are expensive and difficult to retrofit. The bubble cap 358 can achieve separations of around 90 %, which in general will be satisfactory.

The hot, at least partially regenerated, catalyst is collected as a dense phase fluidized bed 75 in the base of the existing regenerator shell 24. Additional regeneration gas may be added, by air distributor means 375. It is also conventional to add some fluffing air, to maintain the dense bed in a fluidized state.

It is beneficial to recycle some hot regenerated catalyst from the bubbling dense bed to the fast fluidized bed region 352. Catalyst recycle is usually needed to "fire up" the coke combustor, and achieve the high temperatures needed in the coke combustor for efficient coke combustion and to promote afterburning in the dilute phase transport riser 356. Such recycle preferably comes from the dipleg 404 of a primary cyclone 400. Flue gas is removed from the unit via line 402, while catalyst is discharged from the cyclone into funnel collector 407, from which any catalyst not recycled simply overflows into bed 75. This arrangement is beneficial for two reasons, high

temperature and head, each of which will be briefly reviewed.

Hot regenerated catalyst recovered from flue gas in a cyclone will in most instances be at the very highest temperature in the regenerator. When the unit is being pushed hardest, or is upset, there will be more coke combustion shifted into the dense bed region 75 and less in the coke combustor. This will increase the temperature in bed 75, and in the dilute phase region above it. When the unit is pushed further, with a significant amount of afterburning occurring in the dilute phase region above bed 75, the catalyst recovered from the cyclones can be 28 to 56°C (50 to 100°F) hotter than bed 75. This means that if too much coke burning is shifted to dense bed 75, and to the dilute phase region 70, the unit automatically produces hotter catalyst for recycle to the coke combustor. This leads to a hotter coke combustor, and will, if sufficient air is added, lead to high coke burning rates in the coke combustor. In this way the unit is somewhat self adjusting.

Catalyst head is important for moving catalyst from the dense bed into the coke combustor reliably and controllably. This is difficult because the coke combustor has an elevation close to that of the dense bed 75, and in some units may have a higher elevation than the dense bed. It is difficult and expensive to move catalyst from a lower elevation to a higher elevation. Use of the natural dynamics of an FCC unit, where much of the catalyst inventory is carried or entrained up into the regenerator cyclones, allows some useful work to be performed by this elevated catalyst. The FCC regenerator entrains or sweeps up into the dilute phase region 70, and through the cyclones 100, all of the catalyst inventory in the regenerator every 5-15 minutes in many units. The cyclones therefore provide a source of catalyst with a large head

available to permit controlled transfer of hot regenerated catalyst from the cyclone dipleg to the fast fluidized bed region, with flow control achieved via slide valve 405. Regenerated catalyst is then charged to the FFB region 352 via line 409.

In some cases it will be possible to reduce or eliminate, the recycle of regenerated catalyst to the FFB region 352. This is because of the countercurrent heat exchange possible between relatively cool spent catalyst in the stripper standpipe 26 and the hotter catalyst in FFB region 352 and in the dilute phase transport riser 356. Use of conductive refractory linings, or other materials of construction which assist heat transfer between dense and dilute catalyst phases, will augment heat transfer between stripped catalyst and regenerated catalyst and flue gas.

Figure 6 shows several modifications of the third example, in which heat is removed from around the regenerator 24. In most units, only one or two methods of heat removal would be practiced, but three methods of heat removal are shown, and all three could be practiced simultaneously. All three modes of heat removal are in addition to the catalyst cooler 28 shown in Figure 1 (prior art).

In the the example shown in Figure 6, heat can be removed from the catalyst recycled to the FFB region from either the dense bed or from a cyclone dipleg. When the highest grade steam is needed, it is beneficial to remove heat from catalyst recycled from the hottest place in the regenerator, which will usually be the catalyst recovered from cyclone diplegs. Thus, as shown in Figure 6, catalyst is withdrawn from dipleg 501, passed via line 503 into heat exchange means 505, usually a conventional catalyst cooler. A heat exchange fluid, usually water or low grade steam, will be added to the cooler 505 via line 507 and removed via line 508. Resulting cooled catalyst can be

charged to the FFB region of coke combustor 550 via line 510.

Heat removal from a catalyst recycle line associated with the dense bed region 75 can also be practiced. In this embodiment, a line 515 removes catalyst from the dense phase region 75 of the FFB, and passes it through heat exchange means 519, where a cool incoming fluid in line 527 is heated and then removed via line 529. The cooled catalyst will then be charged via line 530 to the FFB region.

A preferred method of heat removal is to install a heat removal means 535 in the transfer line 32 removing catalyst from the dense bed region and returning it to the riser reactor 6. A coolant fluid can be added via line 537, heat exchanged, and removed via line 539. This means that a much cooler catalyst will be used in the reactor, so higher catalyst:oil ratios can be achieved in the unit, with consequent increases in conversion and gasoline yields.

FCC FEED

Any conventional FCC feed can be used. The process of the present invention is especially useful for processing difficult charge stocks, those with high levels of CCR material, exceeding 2, 3, 5 and even 10 wt % CCR.

The feeds may range from the typical, such as petroleum distillates or residual stocks, either virgin or partially refined, to the atypical, such as coal oils and shale oils. The feed frequently will contain recycled hydrocarbons, such as light and heavy cycle oils which have already been subjected to cracking.

Preferred feeds are gas oils, vacuum gas oils, atmospheric resids, and vacuum resids, and mixtures thereof. The present invention is most useful with feeds having an initial boiling point above about 343°C (650°F).

The most uplift in value of the feed will occur when a significant portion of the feed has a boiling point above about 540dC (1000°F), or is considered non-distillable, and when one or more heat removal means are provided in the regenerator, as shown in Figure 2 or in Figure 3.

FCC CATALYST

Any commercially available FCC catalyst may be used. The catalyst can be 100% amorphous, but preferably includes some zeolite in a porous refractory matrix such as silica-alumina, clay, or the like. The zeolite is usually 5-40 wt.% of the catalyst, with the rest being matrix. Conventional zeolites include X and Y zeolites, with ultra stable, or relatively high silica Y zeolites being preferred. Dealuminized Y (DEAL Y) and ultrahydrophobic Y (UHP Y) zeolites may be used. The zeolites may be stabilized with Rare Earths, e.g., 0.1 to 10 Wt % RE.

Relatively high silica zeolite containing catalysts are preferred for use in the present invention. They withstand the high temperatures usually associated with complete combustion of CO to CO2 within the FCC regenerator.

The catalyst inventory may also contain one or more additives, either present as separate additive particles, or mixed in with each particle of the cracking catalyst. Additives can be added to enhance octane (shape selective zeolites, i.e., those having a Constraint Index of 1-12, preferably ZSM-5), adsorb SOX (alumina), remove Ni and V (Mg and Ca oxides).

CRACKING REACTOR/STRIPPER/REGENERATOR

The FCC reactor, stripper and regenerator shell 24, per se, are conventional, and are available from the M.W. Kellogg Company.

The modifications needed to add the coke combustor within, or built partially into, the base of the existing regenerator shell 24, and the optional radial

FFB region shown in Figure 2 are well within the skill of the art.

REGENERATOR CONDITIONS

In the coke combustor, the air admission rate, and the cross-sectional area available for flow, and catalyst addition and catalyst recycle, if any, are adjusted to maintain much or all of the catalyst in at least a turbulent fluidized condition, and preferably in a "fast fluidized condition", characterized by intense agitation, relatively small bubbles, and rapid coke combustion. In terms of superficial vapor velocity and typical FCC catalyst sizes, this means the vapor velocity should exceed 1 m/second (3.5 feet/second), and preferably should be 1.2-4.6 m/second (4-15 feet/second), most preferably 1.2-3 m/second (4-10 feet/second). The catalyst density in a majority of the volume in the coke combustor will be less than 0.6 gm/cc (35 pounds/cubic foot), and preferably less than 0.5 gm/cc (30 pounds/cubic foot), and ideally about 0.4 gm/cc (25 pounds/cubic foot).

The densities and superficial vapor velocities discussed herein presume that the unit operates at a pressure where the vast majority of FCC units operate, namely 270-380 kPa (25-40 psig). Changes in pressure change the superficial vapor velocity needed to maintain, e.g., a fast fluidized bed or a bubbling dense bed. It is easy to calculate the superficial vapor velocity needed to support a given type of fluidization, and the bed density expected at those conditions. In general, an increase in pressure will decrease the superficial vapor velocity needed to achieve a fast fluidized bed.

Immersion of the coke combustor within the bubbling dense bed permits a reduction or elimination of catalyst recycle to the dense bed.

It is preferred to add enough combustion air to the coke combustor to remove 20 to 90 wt% of the coke,

more preferably from 50 to 90 wt% of the coke. Preferably from 5 to 60 wt%, and most preferably 10 to 40 wt%, of the coke combustion occurs in the bubbling bed 75.

It is preferred to add enough combustion air to the bubbling dense bed to remove the remainder of the coke necessary to produce regenerated catalyst with the desired coke level, typically less than 0.1 wt %, and preferably less than 0.05 wt %.

BENEFITS OF FFB COKE COMBUSTION

The process of the present invention achieves several important objectives in the shell of an existing regenerator. Among the objectives are increased coke burning capacity, reduced NOx emissions, and reduced catalyst deactivation. Each will be briefly reviewed.

Increased coke burning capacity can be achieved because each square foot of the old bubbling bed regenerator can be used as productively as before, while the coke combustor region(s) burns two to three times as much coke per square foot of cross sectional area as compared to a bubbling bed regenerator. In the embodiments which include means for separating the catalyst and flue gas discharged from the coke combustor, there will be a net reduction in catalyst traffic in the dilute phase. Moreover, to the extent that catalyst is decoked in the coke combustor, the air feed rate to the bubbling dense bed will be reduced, and this will in turn reduce catalyst entrainment from the dense bed into the dilute phase region.

Reduced NOx emissions can be achieved because most of nitrogen compounds are burned under relatively mild, perhaps even partially reducing conditions in the coke combustor. The presence of a reducing atmosphere, and the presence of carbon, both of which occur more in the coke combustor than anywhere else in the regenerator,

tend to suppress formation of NOx, so that large amounts of coke combustion can be achieved without inordinate amounts of NOx being formed.

Improved catalyst stability is obtained by steaming the catalyst less. More than 90% of the "fast coke" or hydrogen rich coke is removed in the coke combustor, under fast fluidized bed regeneration conditions. The complete regeneration of the catalyst, and removal of the "hard coke", and the highest temperatures, and the most oxidizing conditions, can be left to the bubbling fluidized bed. This staged combustion allows most of the water of combustion to be formed and rapidly removed, in the flue gas from the coke combustor, allowing drier regeneration of catalyst in the downstream regions, e.g., the bubbling dense bed. There will still be some thermal deactivation of the catalyst in the bubbling dense bed and as the catalyst is carried into the dilute phase region above the bubbling dense bed regenerator. Thus, if the dilute phase region above the coke combustor and the dense bed is not partitioned, water of combustion formed in the coke combustor can increase the steam partial pressure in the dilute phase region above the dense bed.

Using the coke combustor to burn most of the coke on the stripped catalyst also allows the amount of excess air to be greatly reduced. Typical bubbling dense bed units operate with around 2 % oxygen in the flue gas. Use of an internal coke combustor immersed in a bubbling dense bed, and receiving catalyst recycle from the bubbling dense bed, allows much less excess air to be used, with operation with 0.3 - 0.8 % oxygen in the flue gas being possible.

CO COMBUSTION PROMOTER

Use of a CO combustion promoter in the regenerator is not essential for the practice of the present invention, however, it is preferred. These materials are well-known.

U.S. 4,072,600 and U.S. 4,235,754 disclose operation of an FCC regenerator with minute quantities of a CO combustion promoter. From 0.01 to 100 ppm Pt metal or enough other metal to give the same CO oxidation, may be used with good results. Very good results are obtained with as little as 0.1 to 10 wt. ppm platinum present on the catalyst in the unit.

FCC REACTOR CONDITIONS

Conventional riser cracking conditions may be used. Typical riser cracking reaction conditions include catalyst/oil ratios of 0.5:1 to 15:1 and preferably 3:1 to 8:1, and a catalyst contact time of 0.1 to 50 seconds, and preferably 0.5 to 5 seconds, and most preferably 0.75 to 2 seconds, and riser top temperatures of 480 to 565°C (900 to 1050°F).

CLAIMS

1. A fluidized catalytic cracking process wherein a heavy hydrocarbon feed is cracked to lighter products comprising the steps of:

catalytically cracking said feed in a riser reactor by mixing the feed in the base of the reactor with a source of hot regenerated catalytic cracking catalyst withdrawn from a catalyst regenerator, and cracking said feed in said riser reactor to produce catalytically cracked products and spent catalyst which are discharged from the top of the riser into a catalyst disengaging zone;

separating cracked products from spent catalyst in said catalyst disengaging zone to produce a cracked product vapor phase which is recovered as a product and a spent catalyst phase which is discharged from said disengaging zone into a catalyst stripping zone contiguous with and beneath said disengaging zone;

stripping said spent catalyst with a stripping gas in said stripping zone to produce a stripper vapor comprising cracked products and stripped catalyst, which is discharged into a vertical standpipe beneath said stripping zone;

discharging stripped catalyst from said standpipe into a catalyst regenerator and regenerating said stripped catalyst in said regenerator, said regenerator comprising a first regeneration zone which is located beneath said stripping zone and which comprises a single dense phase bubbling fluidized bed of catalyst to which an oxygen containing regeneration gas is added and from which hot regenerated catalyst is withdrawn and recycled to said riser reactor, characterized in that:

said regenerator comprises a second, coke combustion zone which is at least partially immersed in the bubbling fluidized bed of the first regeneration

zone and into which stripped catalyst is discharged from said standpipe;

an oxygen-containing gas is added to said coke combustion zone partially to regenerate the stripped catalyst and maintain a majority of the catalyst in a state of turbulent or fast fluidization; and

the partially regenerated catalyst and flue gas is discharged from said coke combustor to a dilute phase region of the first regeneration zone above the bubbling fluidized bed and at least part of the catalyst is separated from the flue gas and collected in said bubbling fluidized bed for further regeneration.

2. The process of claim 1 wherein separation of the flue gas and partially regenerated catalyst is effected by a deflector cap which extends over the upper end of the coke combustor vessel.

3. The process of claim 1 wherein separation of the flue gas and partially regenerated catalyst is effected by a cyclone separator.

4. The process of claim 1 wherein the coke combustor comprises a vertical, generally cylindrical vessel connected at its upper end to the dilute phase region of the first regeneration zone.

5. The process of claim 1 wherein the coke combustor comprises a vessel having a base region and an upper region which is of reduced cross section compared to the base region and which is connected to the dilute phase region of the first regeneration zone.

6. The process of claim 5 wherein the base region of the coke combustor vessel opens into the bubbling fluidized bed.

7. The process of claim 5 wherein the base region of the coke combustor vessel is closed and extends below the bubbling fluidized bed.

8. The process of claim 1 wherein catalyst is withdrawn from said bubbling dense bed and contacted

with additional regeneration gas in a third regeneration zone comprising a fast fluidized bed region and resulting regenerated catalyst is returned to said bubbling dense bed.

9. The process of claim 1 wherein catalyst from said bubbling dense bed is recycled to said coke combustor.

10. The process of claim 9 wherein a heat exchange means cools catalyst recycled from said bubbling dense bed to said coke combustor.

11. The process of claim 1 wherein 5 to 50 % of the coke on spent catalyst is burnt in the bubbling fluidized bed and 20 to 90% of the coke on spent catalyst is burnt in the coke combustor.

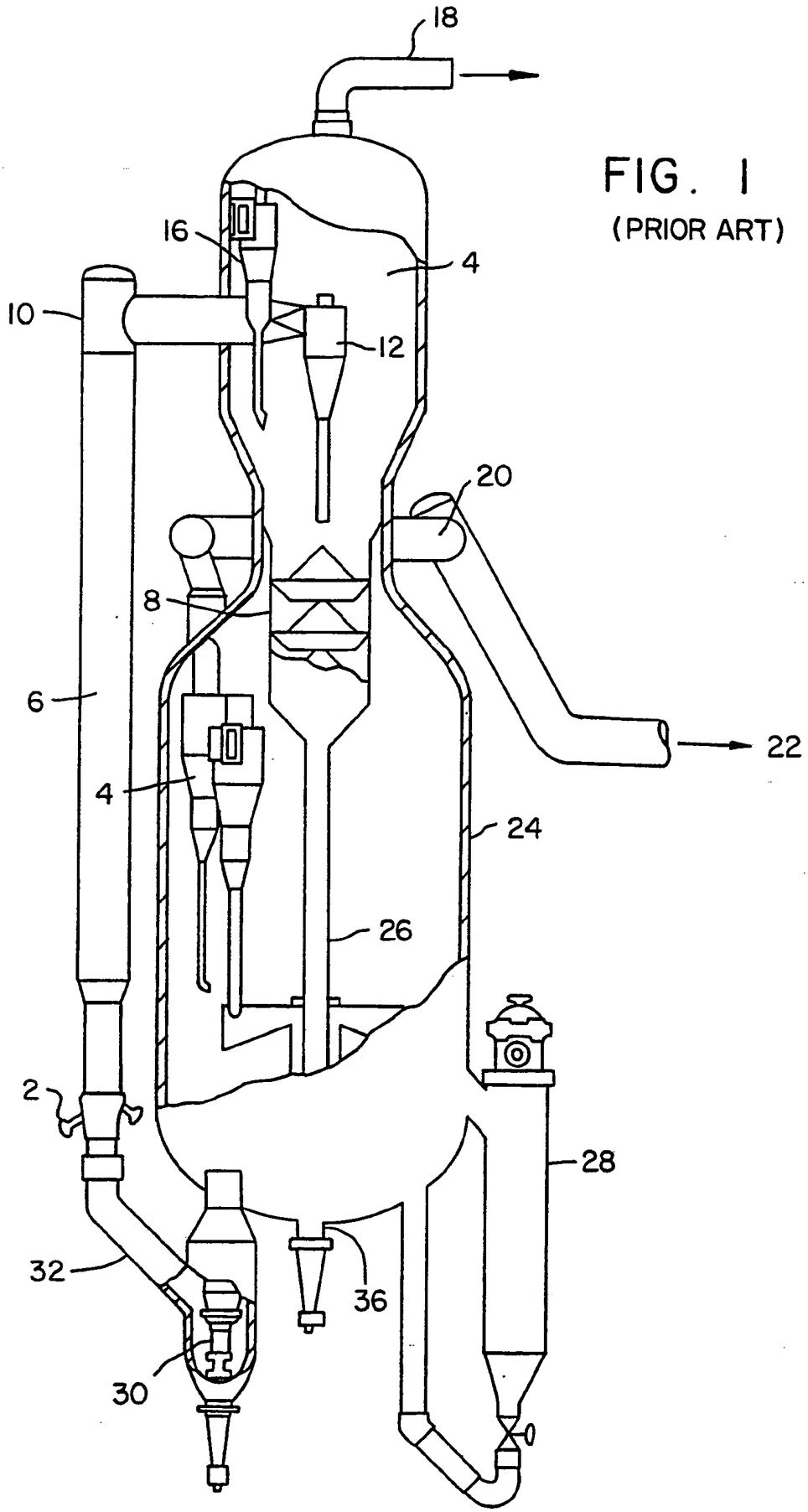

FIG. 1
(PRIOR ART)

FIG. 2

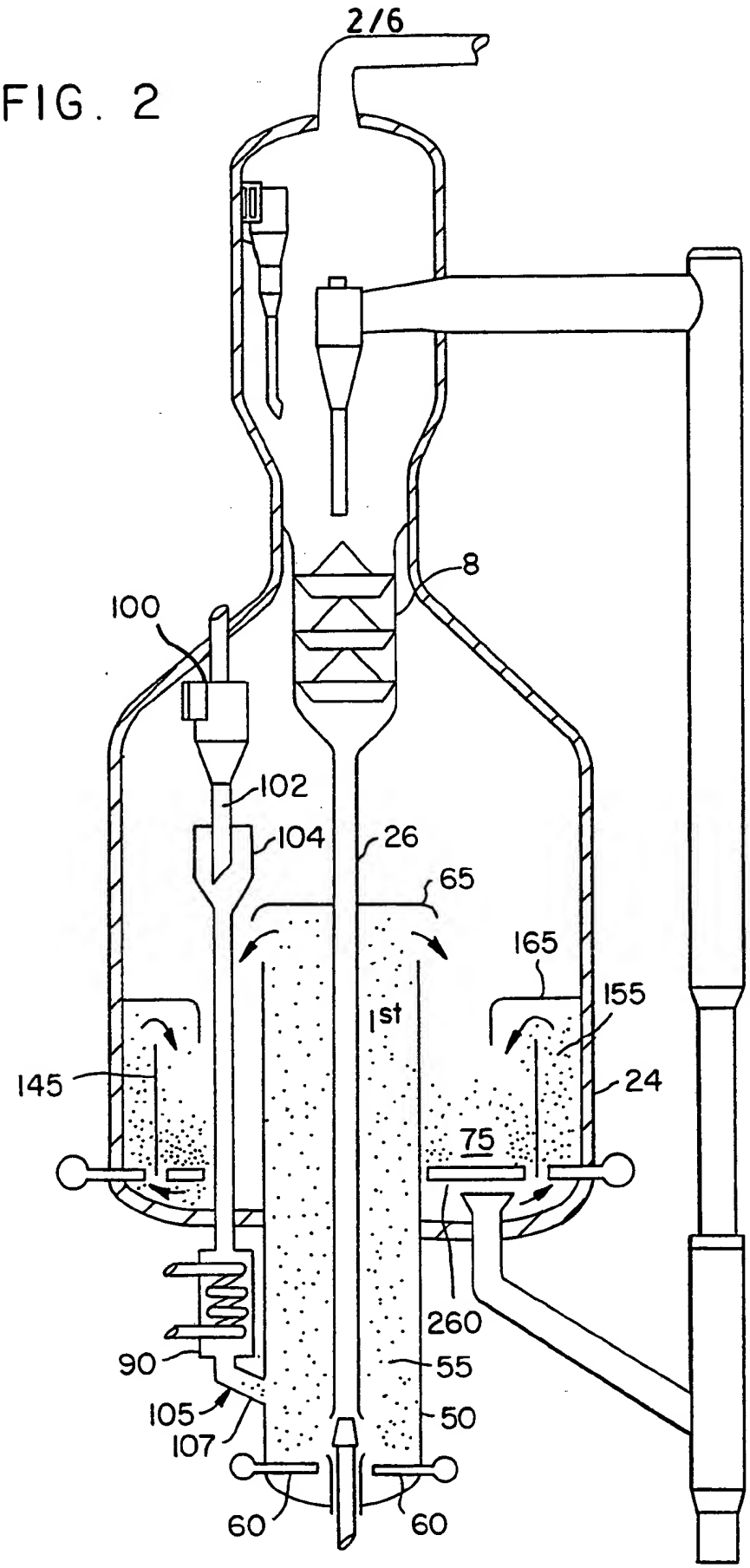
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FIG. 3

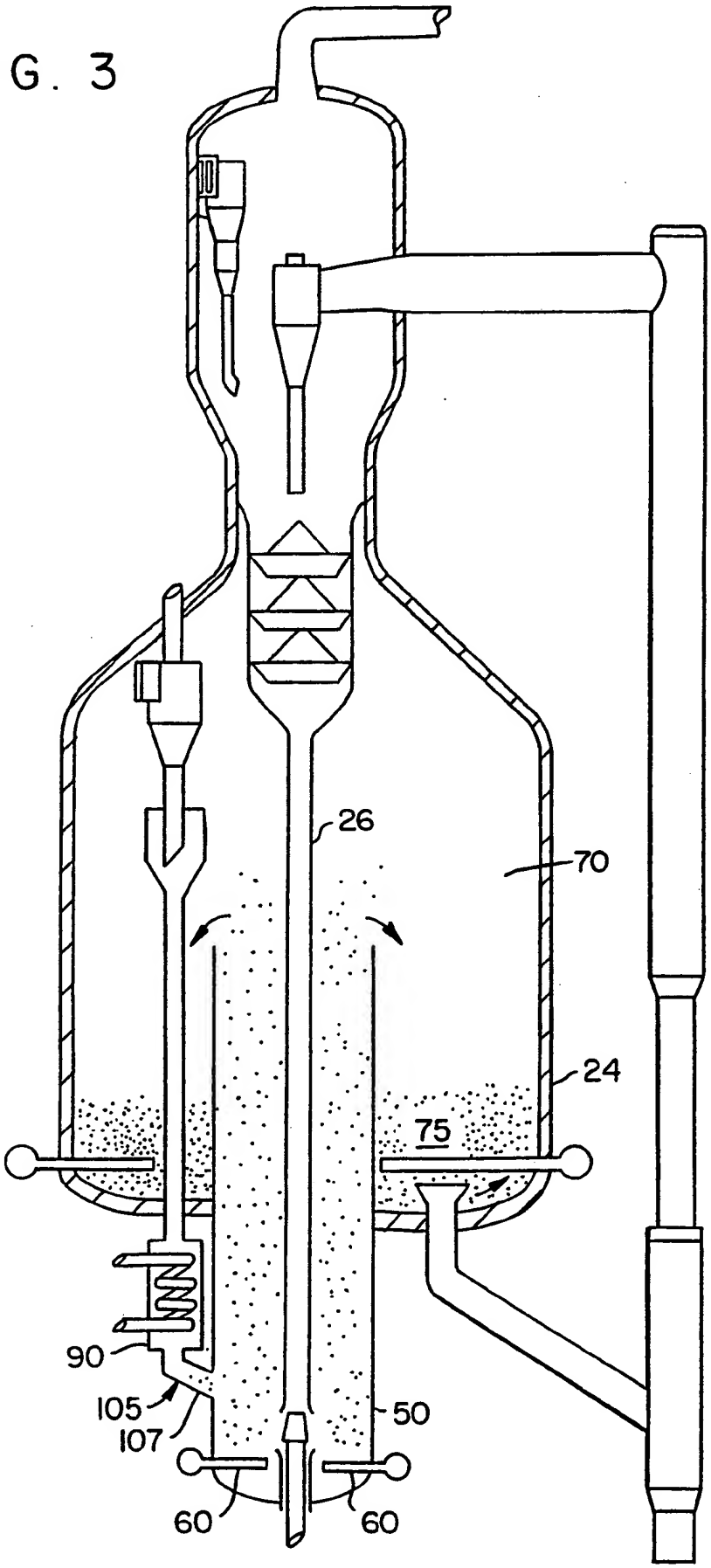

FIG. 4

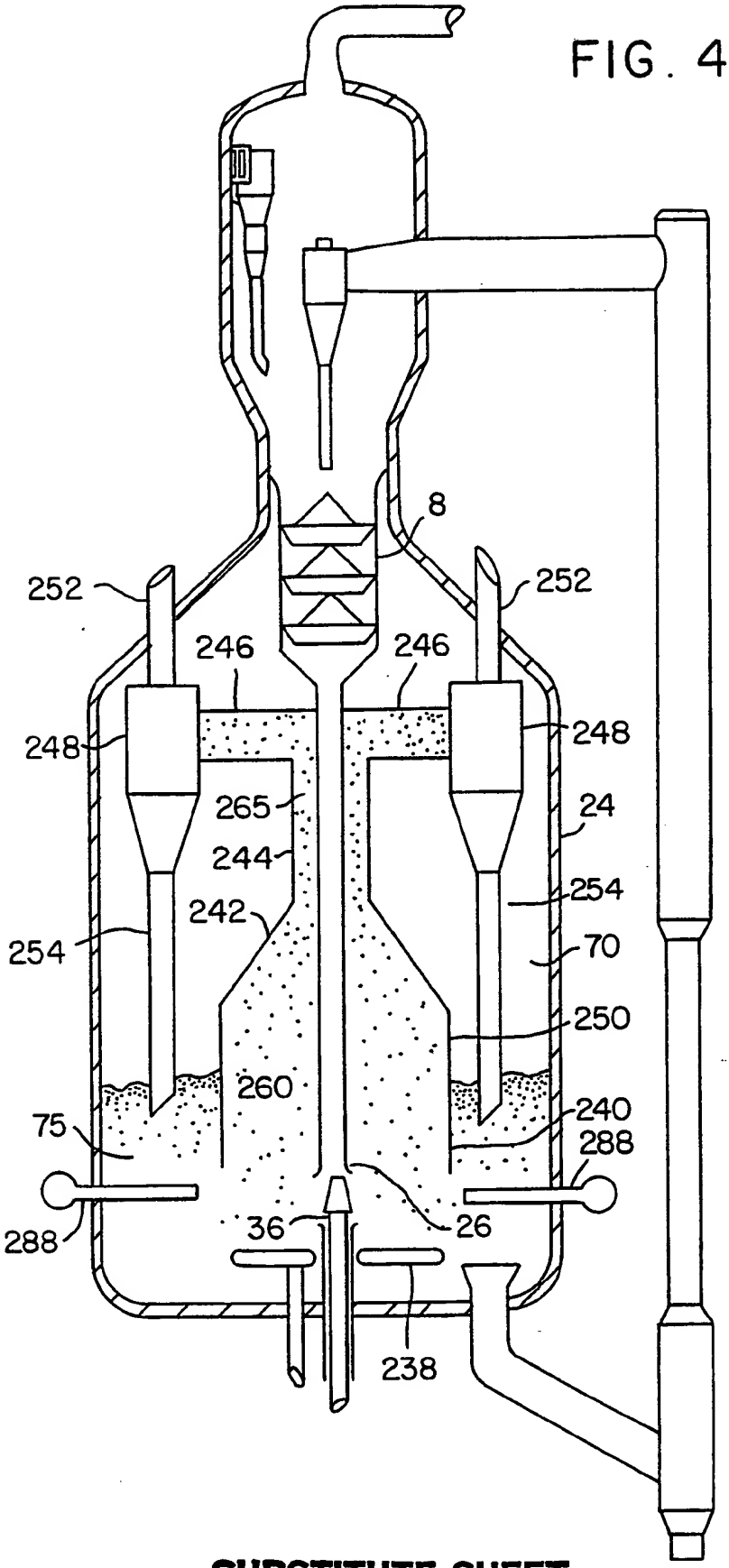
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FIG. 5

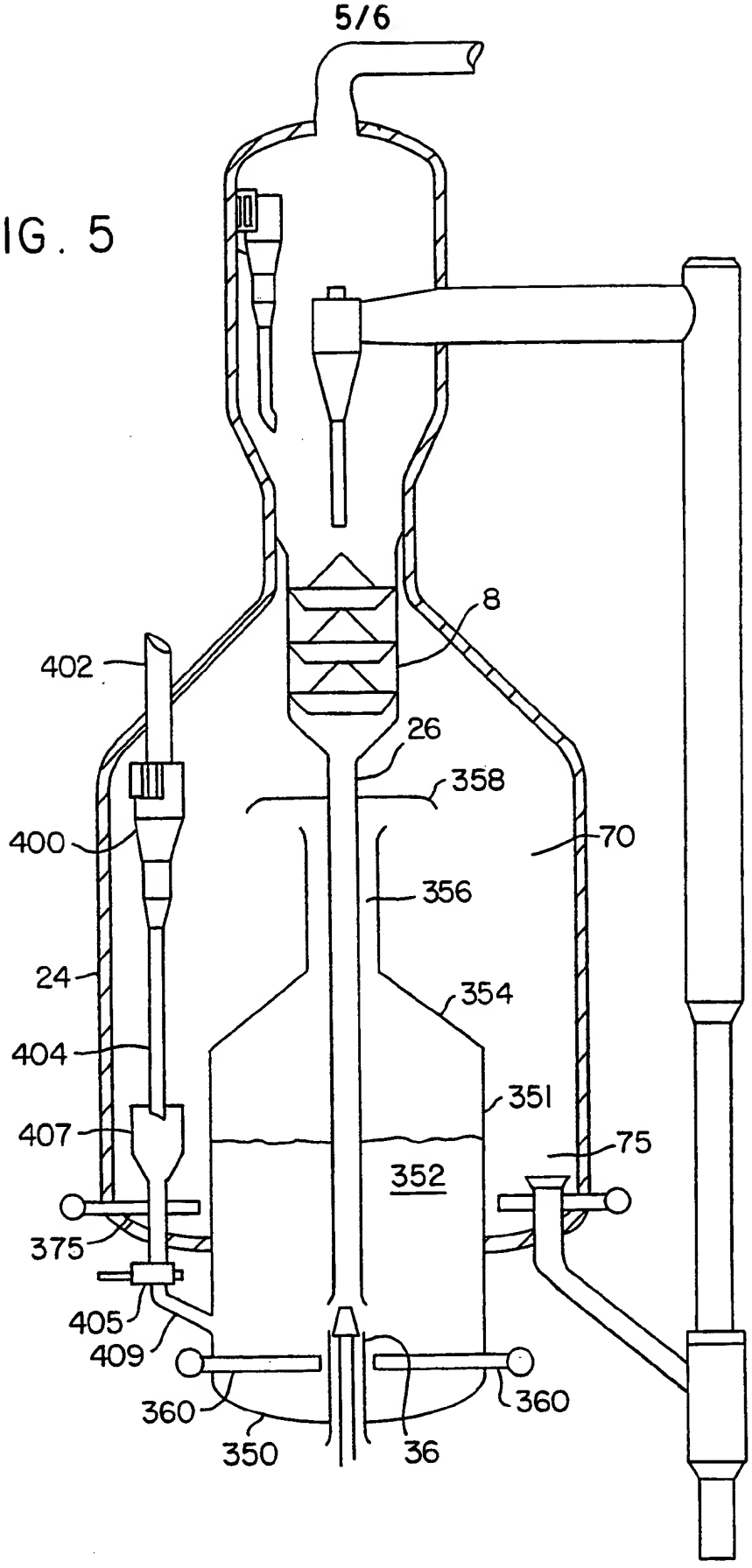
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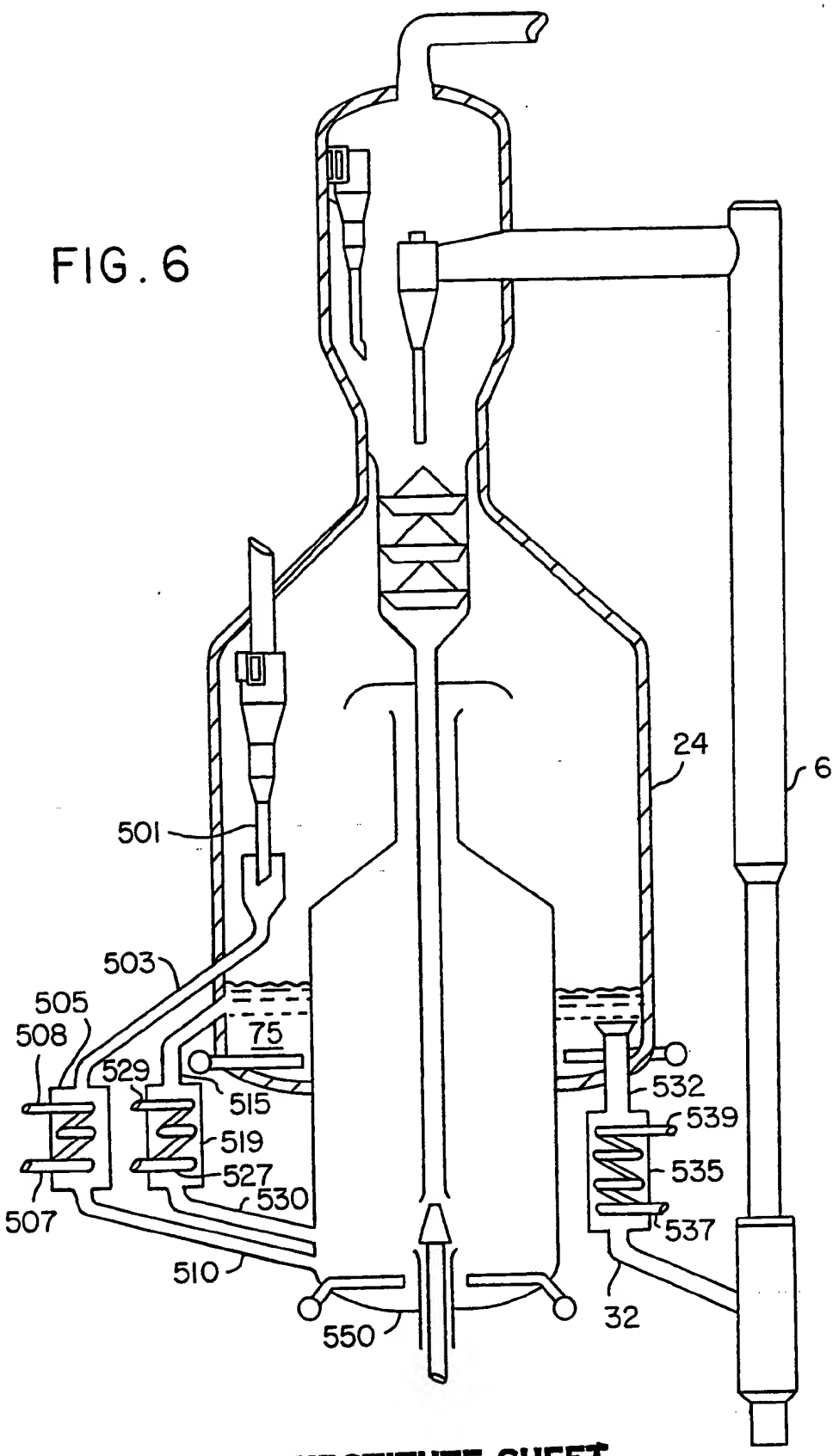
FIG. 6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INTERNATIONAL SEARCH REPORT

International Application No. PCT/US91/04831

I. CLASSIFICATION OF SUBJECT MATTER (if several classification symbols apply, indicate all) [6]

According to International Patent Classification (IPC) or to both National Classification and IPC

IPC(5): C10G 35/10; B01J 20/34

U.S. CL: 208/113, 158, 159, 160, 164

II. FIELDS SEARCHED

Minimum Documentation Searched [7]

Classification System	Classification Symbols
U. S.	208/113, 158, 159, 160, 164

Documentation Searched other than Minimum Documentation to the Extent that such Documents are Included in the Fields Searched [8]

III. DOCUMENTS CONSIDERED TO BE RELEVANT [9]

Category *	Citation of Document, [11] with indication, where appropriate, of the relevant passages [12]	Relevant to Claim No. [13]
A	US, A, 4,812,430, MARCH 1989 (CHILD) See Column 6, Line 16 to 51.	1-11
A	US, A, 4,820,404, 11 APRIL 1989 (OWEN) See Column 11, Line 25 to 63.	1-11
A	US, A, 4,118,338, OCTOBER 1978 (GROSS ET. AL.) See column 9, lines 1-43.	1-11

* Special categories of cited documents: [10]

"A" document defining the general state of the art which is not considered to be of particular relevance

"E" earlier document but published on or after the international filing date

"L" document which may throw doubts on priority claim(s) or which is cited to establish the publication date of another citation or other special reason (as specified)

"O" document referring to an oral disclosure, use, exhibition or other means

"P" document published prior to the international filing date but later than the priority date claimed

"T" later document published after the international filing date or priority date and not in conflict with the application but cited to understand the principle or theory underlying the invention

"X" document of particular relevance; the claimed invention cannot be considered novel or cannot be considered to involve an inventive step

"Y" document of particular relevance; the claimed invention cannot be considered to involve an inventive step when the document is combined with one or more other such documents, such combination being obvious to a person skilled in the art.

"&" document member of the same patent family

IV. CERTIFICATION

Date of the Actual Completion of the International Search	Date of Mailing of this International Search Report
23 MARCH 1992	20 APR 1992
International Searching Authority	**Signature of Authorized Officer**
ISA/US	HELANE MYERS for

Form PCT/ISA/210 (second sheet) (Rev.11-87)